PLANTWIDE CONTROL:
The search for the self-optimizing control structure

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QUESTION:

- Why do we control all these variables for which there are no *a priori* specifications? (temperatures, pressures, internal compositions, . . .)

The Trondheim **Plantwide control group** :

- **Audun Faanes** (pH-neutralization; control structure design)
- **Marius Govatsmark** (Refinery; HDA plant)
- **Ivar Halvorsen** (Petlyuk distillation; self-optimizing control)
- **Truls Larsson** (methanol plants; control structure design)
- **Tore Lid** (Refinery)
- **Sigurd Skogestad**

associated (earlier) members

- John Morud (Sintef Kjemi)
- Kjetil Havre (ABB)
- Bjørn Glemmestad (Borealis)

plus

- Morten Hovd, *Department of Engineering Cybernetics, NTNU*
- Stig Strand, *Statoil*

OUTLINE

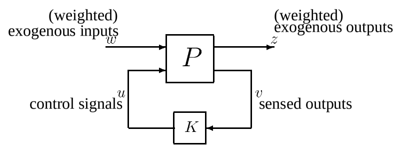
1. Introduction: Theory and practice
2. Plantwide control and control structure design
3. Controlled variables
4. Self-optimizing control
5. Previous work on self-optimizing control
6. Procedure for selecting controlled variables
7. Distillation case study
8. Optimal operation: An ethical dilemma?
9. Conclusion

Written material:

- *Plantwide control: The search for the self-optimizing control structure*, S. Skogestad (prepared for publication in J.Proc.Control). Avaialable at: `http://www.chembio.ntnu.no/users/skoge/publications/2000/self1.pdf`
- Shorter version: *Self-optimizing control: The missing link berween steady-state optimization and control*, S. Skogestad, Symposium Process Systems Engineering PSE'2000, Keystone, Colorado, July 2000. Avaialable at: `http://www.chembio.ntnu.no/users/skoge/publications/2000/self1`$_{pse}$*2000.pdf*
- *Multivariable Feedback Control*, S. Skogestad and I. Postlethwaite, Wiley, 1996
  - Chapter 5: Limations on performance in SISO systems (controllability)
  - Chapter 6: Limations on performance in MIMO systems (more controllability)
  - Chapter 10: Control structure design (plantwide control)
- *A review of plantwide control*, S. Skogestad and T. Larsson, (Internal report, Aug. 1998). Available at: `http://www.chembio.ntnu.no/users/skoge/publications/1998/plantwide review1.pdf`
- Ph.D. thesises of Morud (1995), Glemmestad (1997) and Havre (1998).

EXISTING CONTROL THEORY

General controller design formulation (Ph.D.level)



- w: Disturbances (d) and setpoints (r)
- v: Measurements (y_m, d_m) and setpoints (r)
- u: Manipulated inputs (u)
- z: Control error, $y - r$
- Find a controller K which based on the information in v, generates a control signal u which counteracts the influence of w on z, thereby minimizing the closed-loop norm from w to z.

PRACTICE
Typical control hierarchy

PRACTICE
Typical base level control structure
PID level

Relevant questions in practice

1. Is the plant easy or difficult to control? (*controllability analysis*)?
2. How should the plant be controlled (*plantwide control*) ? (THIS TALK)

Alan Foss (“Critique of chemical process control theory”, AIChE Journal, 1973):

> *The central issue to be resolved ... is the determination of control system structure.*
>
> *Which variables should be measured, which inputs should be manipulated and which links should be made between the two sets?*
>
> *The gap is present indeed, but contrary to the views of many, it is the theoretician who must close it.*

PLANTWIDE CONTROL (Control structure design)

The *control philosphy* for the overall plant with emphasis on the *structural decisions*:

- Which “boxes” (controllers; decision makers) do we have and what information (signals) are send between them

Not:

- The inside of the boxes (design and tuning of all the controllers)

CONTROL STRUCTURE DESIGN TASKS

1. *Selection of controlled outputs* (a set of variables which are to be controlled to achieve a set of specific objectives)
2. *Selection of manipulations and measurements* (sets of variables which can be manipulated and measured for control purposes)
3. *Selection of* **control configuration** (a structure interconnecting measurements/commands and manipulated variables)
4. *Selection of controller type* (control law specification, e.g., PID, decoupler, LQG, etc.).

Tasks 1 and 2 combined: **input/output selection**
Task 3 (configuration): **input/output pairing**

Approach Control structure design:

- Top-down consideration of control objectives and available degrees of freedom to meet these (tasks 1 and 2)
- Bottom-up design of the control system, starting with stabilization (tasks 3, 4, 5).

Controlled variables

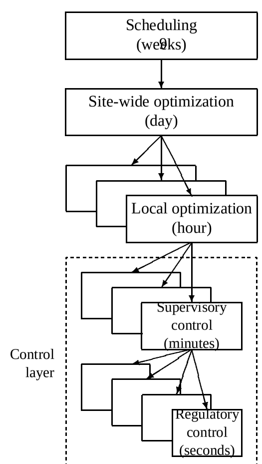


Figure .1: Typical control hierarchy in a chemical plant.

The setpoints of the controlled variables (c_s) are the (internal) variables that link two layers in a control hierarchy, whereby the upper layer computes the value of c_s to be implemented by the lower layer.

TASK 1: Selection of controlled variables

Two distinct questions:

1. **Optimization:** What are the optimal values (e.g. $c_{\rm opt}$)?
2. **Implementation:** What should be the controlled variables c?
   (includes open-loop by selecting $c = u$)

First question: A lot of theory.

BUT second question: Almost no theory. Decisions mostly made on experience and intuition.

Optimization

Minimize cost J with respect to the degrees of freedom u

$$\min_u J_u(u,d) \tag{1}$$

subject to constraints

$$g(u,d) \le 0 \tag{2}$$

Gives $u_{\mathrm{opt}}(d)$, $c_{\mathrm{opt}}(d)$, etc.

Degrees of freedom for optimization

$$N_{\mathrm{opt}} = N_u = N_m - N_0$$

"Free" degrees of freedom after satisfying constraints

$$N_{\mathrm{opt,free}} = N_{\mathrm{opt}} - N_{\mathrm{active}}$$

Example: Cake baking

Goal (purpose): Well-baked inside and nice outside

Degree of freedom: Heat input $u = Q$

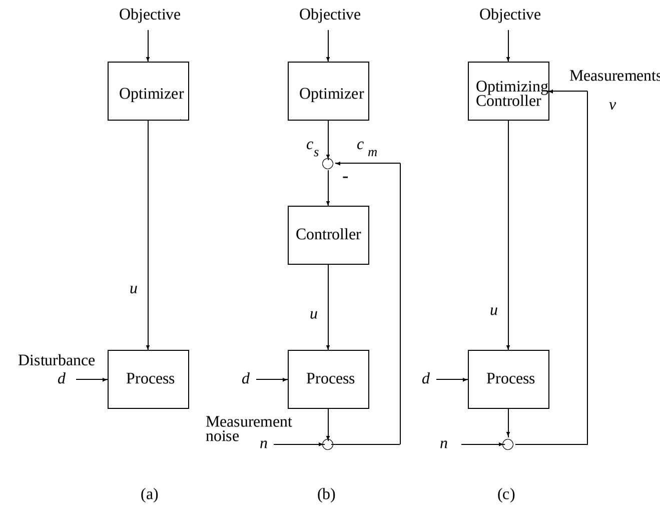
Implementation:

1. *Open-loop implementation:* Constant heat input Q
   - Problem: Sensitive to uncertainty and optimal value depends strongly on size of oven

2. *Closed-loop implementation:* Constant temperature
   - $c =$ oven temperature
   - "Optimizer": Cook book (look-up table)

   Used in practice. Insensitive to changes ("self-optimizing").

Implementation



(a) "Open-loop". Constant $u_s = u_{\mathrm{opt}}(d^*)$

(b) "Closed-loop". Constant $c_s = c_{\mathrm{opt}}(d^*)$

(c) "Integrated optimization and control". Reoptimize $u = u_{\mathrm{opt}}(d)$.

Self-optimizing control

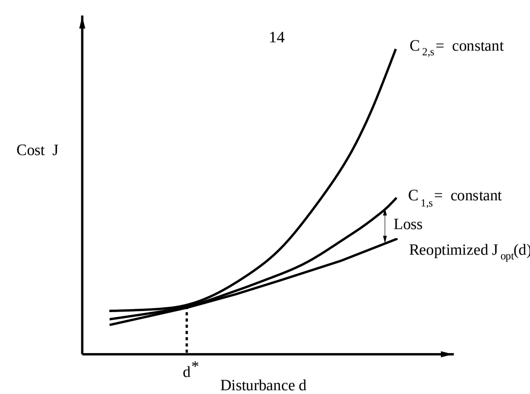


Figure .2: Loss imposed by keeping constant setpoint for the controlled variable

Self-optimizing control is when we can achieve an acceptable loss with constant setpoint values for the controlled variables (without the need to reoptimize when disturbances occur)

Effect of implementation error

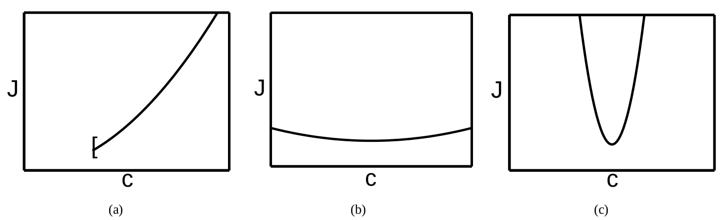


Figure .3: Implementing the controlled variable

(a) Implementation easy: Active constraint control

(b) Implementation easy: Insensitive to error in c

(c) Implementation difficult: Look for another controlled variable]

Requirements for controlled variables

- Skogestad and Postlethwaite (1996; Chapter 10).

The selected set of variables c should:

1. Have a small variation in optimal setpoints
2. Be easy to control accurately (small implementation error)
3. Should be sensitive to the manipulated inputs (u), i.e. have a large range. (Equivalently, the optimum should be flat with respect to c).
4. Be independent (not closely correlated)

The four requirements may be combined into the **singular value rule**:

Prefer a set of controlled variables with a large minimum singular value of the scaled gain matrix, $\underline{\sigma}(G(0))$.

Here

- $\Delta c = G\Delta u$ (linearized model)
- Variables c (and thus G) are scaled such that $\|c - c_{\rm opt}(d)\| \approx 1$ (takes care of requirements 1 and 2)

Previous work on self-optimzing control

- Morari, Stephanopoulos and Arkun (1980):

  "Translate ecomomic objectives into control objectives"

  "Find a function c of process variables, which when held constant leads automatically to the optimal adjustment of the manipulated variables"
- Shinnar (1981); Arbel, Rinard and Shinnar (1996):

  "Dominant variables" (= natural choice for controlled variables c)
- Luyben (1988):

  "Eigenstructure' (= the best self-regulationg and self-optimizing structure)
- Fisher, Dohert and Douglas (1988, part 3): HDA example
- Rijnsdorp (1991):

  "on-line algorithms for adjusting the degrees of freedom for optimization'
- Narraway and Perkins (1991, 1993, 1994):

  Base control structure selection on process economics
- Marlin and Hrymak (1997): Implementation of optimal solution; active constraints
- Zheng, Mahajannam and Douglas (1999): Minimize economic penalty

Selecting controlled variables

Today: Steady-state optimization is performed routinely

Thus: Have the main tools needed for a proper selection of controlled variables:

- Steady-state model
- Information about operational constraints
- Well-defined economic objective (scalar cost function J to be minimized)

Procedure for selecting controlled variables

Step 1: Degree of freedom analysis. Determine the number of degrees of freedom ($N_{\text{opt}} = N_u$) available for optimization, and identify a base set (u) for the degrees of freedom.

Step 2: Cost function and constraints. Define the optimal operation problem by formulating a scalar cost function J to be minimized for optimal operation, and specify the constraints that need to be satisfied.

Step 3: Identify important disturbances (uncertainty).

- Disturbances $d - d^*$ that occur after the optimization
- Implementation errors (d_c) in controlled variables c

Step 4: Optimization.

Find the optimal steady-state operation for various disturbances.

Step 5: Identify candidate controlled variables c.

- Normally implement "active constraints"
- Leaves $N_{\text{opt,free}}$ degrees of freedom:
  - Typically, consider measured variables or simple combinations thereof.
  - Want to use constant $c_s = c_{opt}(d^*)$.
  - Use above requirements, e.g look for variables where $c_{opt}(d)$ is constant

Step 6: Evaluation of loss. Compute the mean value of the loss for alternative sets of controlled variables c. This is done by evaluating the loss

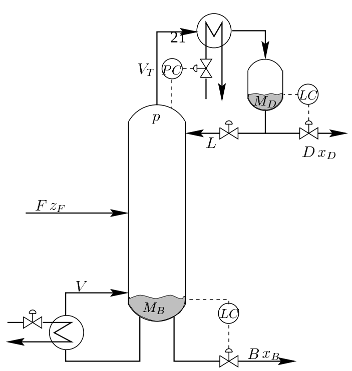
$$L(u,d) = J(u,d) - J(u_{opt}(d),d); \quad u = f_c(c_s + d_c, d) \tag{3}$$

with fixed setpoints c_s for the defined disturbances $d \in \mathcal{D}$ and implementation errors $d_c \in \mathcal{D}_c$.

Step 7: Further analysis and selection.

- Select for further consideration the sets of controlled variables with acceptable loss.
- See if they are adequate with respect to other criteria that may be relevant, such like
  - region of feasibility
  - expected dynamic control performance (input-output controllability)

Case study: Distillation



Propylene-propane splitter

$$\alpha = 1.12$$
$$N = 110; \quad N_F = 39$$
$$z_F = 0.65$$

Step 1: Degree of freedom analysis Given pressure and feedrate

$$N_{\text{opt}} = N_u = 2$$

(e.g. selected as D and V, or L and V)

Step 2a: Cost function

$$-J = P = p_D D \underset{22}{+} p_B B - p_F F - p_V V$$

Prices [$/kmol]

$$p_D = 20; \quad p_B = 10 - 20x_B; \quad p_F = 10$$

Typical profit: $P \approx 4$ $/min

Maximum acceptable loss L: 0.04 $/min = 20 000 $/year

Step 2b: Constraints .

- $x_D \geq 0.995$ (Distillate purity specification)
- $V \leq 20 \quad kmol/min$ (Maximum capacity)
- (and $F = 1 \quad kmol/min$)

Will consider three cases:

Case 1: $p_V \approx 0$ (cheap energy)

Case 2: $p_V = 0.1$ [$/kmol]

Case 3: Same as Case 2, but with feed rate as degree of freedom

Case 1: Cheap energy ($p_V \approx 0$)

Both constraints active

- $x_D = 0.995$ (never overpurify valueable product)
- $V = 20$ kmol/min (to maximize yield of valueable distillate product)

Implementation "trivial": Constant x_D and increase V until achieve max. pressure drop (avoid flooding)

Step 4: Optimization

	x_D	x_B	D/F	L/F	V/F	L/D	P/F
Nominal	0.995	0.040	0.639	15.065	15.704	23.57	4.528
$F = 1.3$	0.995	0.040	0.639	15.065	15.704	23.57	4.528
$z_F = 0.5$	0.995	0.032	0.486	15.202	15.525	31.28	2.978
$z_F = 0.75$	0.995	0.050	0.741	14.543	15.284	19.62	5.620
$q_F = 0.5$	0.995	0.040	0.639	15.133	15.272	23.68	4.571
$x_D = 0.996$	0.996	0.042	1.274	15.594	16.232	24.47	4.443

Nominal values: $F = 1, z_F = 0.65, q_F = 1.0, p_D = 20, p_V = 0.1$

Table 1. Optimal operating point (with maximum profit P/F)

Step 5: Candidate controlled variables .

- Bottom product purity constraint always active: $x_D = 0.995$
- One unconstrained degree of freedom, $N_{\text{opt,free}} = 1$
- From Table 1: Control x_B at 0.04
  BUT: Expensive and control problems
- Six alternative controlled variables are considered

$$x_B;\ D/F;\ L;\ L/F;\ V/F; L/D$$

Case 2: Trade-off between energy and yield ($p_V = 0.1$ [$/kmol])

More detailed analysis needed

Step 3: Disturbances .

d_1: 30% increase in F

d_2: z_F from 0.65 to 0.5

d_3: z_F from 0.65 to 0.75

d_4: Liquid fraction from 1.0 (pure liquid) to 0.5 (50% vaporized)

d_5: Safety margin on x_D: increase purity from 0.995 to 0.996

In addition d_c: 20% implementation errors for c

Step 6: Evaluation of loss

	$x_B =$ 0.04	$D/F =$ 0.639	$L =$ 15.065	$L/F =$ 15.065	$V/F =$ 15.704	$L/D =$ 23.57
Nominal	0	0	0	0	0	0
$F = 1.3$	0	0	**0.514**	0	0	0
$z_F = 0.5$	0.023	inf	0.000	0.000	0.001	**1.096**
$z_F = 0.75$	0.019	**2.530**	0.006	0.006	0.004	**0.129**
$q_F = 0.5$	0.000	0.000	0.001	0.001	0.003	0.000
$x_D = 0.996$	**0.086**	**0.089**	**0.091**	**0.091**	**0.091**	**0.093**
20% impl.error	0.012	inf	**0.119**	**0.119**	**0.127**	**0.130**

inf denotes infeasible operation

20% impl.error: $x_B = 0.048, D/F = 0.766, L = 18.08, L/F = 18.08, V/F = 18.85, L/D = 28.28$

Unacceptable loss (larger than 0.04) shown in bold face

Table 2. Loss [$/min] for distillation case study.

- Loss small with x_B constant
- Also small loss: L/F or V/F constant.
- Poor: D/F (or D) constant
- All insesnitive to feed enthalpy (q_F)
- Relatively poor: L/D
- Large loss with $x_D = 0.996$ (overpurification)

- Include "feedforward" from F for L and V

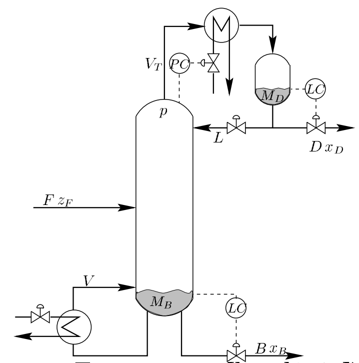
Step 7: Selection of controlled variables . Lowest loss:

$$c_1 = \begin{bmatrix} x_D \\ x_B \end{bmatrix}; \quad c_2 = \begin{bmatrix} x_D \\ L/F \end{bmatrix}; \quad c_3 = \begin{bmatrix} x_D \\ V/F \end{bmatrix};$$

- "Two-point" structure c_1: Difficult (interactive) control problem
- "One-point" c_2 or c_3 preferred

Proposed control system.

- L is used to keep $x_D = 0.995$.
- $V/F = 15.70$ is kept constant (because V may reach constraint).



Case 3: Feedrate as degree of freeedom

- Three degrees of freedom
- Additional constraint

$$F \leq F_{max}$$

Consider "available" feed rate F_{max} as a parameter.

1. **Low available feed rates**; $F_{max} \leq 1.274$ [kmol/min]). (Same as above)

   Proposed control system:

   - L is used to keep $x_D = 0.995$ (active constraint).
   - V is adjusted to keep $V/F = 15.70$ constant
   - F is kept at its maximum available value F_{max} (active constraint).

2. **Intermediate available feed rates**; $1.274 < F_{max} < 1.566$ [kmol/min].

   Proposed control system:

   - L is used to keep $x_D = 0.995$ (active constraint).
   - V is kept at $V_{max} = 20$ (active constraint).
   - F is kept at its maximum available value F_{max} (active constraint).

   As F_{max} is increased from 1.274 to 1.566, the optimal value x_B of increases from 0.04 and 0.09.

3. **Large available feed rates**; $F_{max} > 1.566$ [kmol/min]).

   Proposed control system:

   - L is used to keep $x_D = 0.995$ (active constraint).
   - V is kept at $V_{max} = 20$ (active constraint).
   - F is adjusted to keep $V/F = 12.77$ constant (alternatively, we may keep $F = 1.566$ constant, but using V/F is better if the value of V_{max} changes)

   In this case: $x_B = 0.09$

Optimal operation: An ethical dilemma?

Both design and operation: Maximize profit [$/min]

$$\text{Profit} = \text{Product value} - \text{Feed cost} - \text{Variable costs}$$

For example

$$P = p_D D + p_B B - p_F F - p_V V$$

Design

- F given
- Maximize profit P = maximize value increase P/F [$/kg]

Operation

- Often: F a degree of freedom (with no effect on prices)
- Maximize profit P: Operate at capacity constraint(s) and increase F until zero limiting value increase; $dP/dF = 0$
- Smaller specific value increase P/F than design
- Less effective use of raw materials and energy than design

Question:

- An environmental and ethical dilemma?

A plantwide control design procedure

Bottom up design: (With given controlled and manipulated variables)	**Analysis**
1. REGULATORY CONTROL LAYER. The purpose of this control layer is to enable manual operation of the plant	
Stabilization. Selection of measurements and inputs for stabilization (including slowly drifting modes).	*Pole vectors* Give insight about which measurements and inputs can be used for each unstable mode. Select large elements: Small input energy needed and large noise tolerated.
Local disturbance rejection. Often based on secondary measurements.	*Partially controlled plant* Select secondary measurements (y_2) so that the effect of disturbances on the primary output (y_1) can be handled by the operators.
2. SUPERVISORY CONTROL LAYER. The purpose of this control layer is to control the primary control variables (c).	
Decentralized control. We use a decentralized control structure if the process is noninteracting and the constraints are not changing. Feed-forward control and ratio control may be useful here.	*Controllability analysis for decentralized control* Pair on relative gain array close to identity matrix at crossover frequency, provided not negative at stead-state. Closed loop disturbance gain (CLDG) and performance gain array (PRGA) may be used to analyze the interactions.
Multivariable control Use for interacting process (coordination including feed-forward control). MPC is useful for tracking the active constraints (if the steady-state optimization shows that the active constraints are changing with disturbances).	
3. REAL TIME OPTIMIZATION	
Validation	Nonlinear dynamic simulation

A plantwide control design procedure

Step	Tools (in addition to insight)
Top-down analysis:	
1. CONTROLLED VARIABLES: What is the control objective and which variables should be controlled? Goal: Obtain primary controlled variables ($y_1 = c$)	*Steady-state model and operational objectives* Degree of freedom analysis. Determine the major distrubances. Evaluate the (economic) loss, with constant controlled variables and look for "self-optimizing" control structure.
2. PRODUCTION RATE: Where should the throughput be set? The throughput manipulator cannot be used for other control tasks, thus this choice will have a large effect on the structure of the control system.	The optimal choice may follow from steady-state optimization, but may require continuous reconfiguration (may use MPC to avoid this) The throughput manipulator should have a strong and direct effect on the production rate.

CONCLUSION

QUESTION

- Why do we control all these variables for which there are no *a priori* specifications? (temperatures, pressures, internal compositions, . . .)

ANSWER:

- We have degrees of freedom that need to be specified to achieve optimal operation
- The controlled variables are the links between steady-state economics and process control
- Select the controlled variables to achieve *self-optimizing control*